

SPECIFICATION AMENDMENTS

On page 1, insert above line 1, insert--Priority Claim

The present application claims priority on European Patent Application 02254499.3 filed 26 June 2002.--

On page 1, above line 1, insert--Field of the Invention--

On page 1, above line 6, insert--Background of the Invention--

Paragraph at line 13 of page 1, ending at line 9 of page 2, has been amended as follows:

-- In several recent documents reference is made to abundant gaseous hydrocarbon feedstocks as natural gas and/or associated gas, at remote locations (e.g. in the desert, tropical rain forest) and/or off-shore locations, where no direct use of the gas is possible, usually due to the absence of large human populations and/or the absence of any industry. Transportation of the gas, e.g. through a pipeline or in the form of liquefied natural gas, requires extremely high capital expenditure or is simply not practical., particularly ~~This holds even more~~ in the case of relatively small gas production rates and/or gas fields. Reinjection of the gas (and production at a later moment) is another possibility, however, this will add to the costs of the production, and may, in the case of associated gas, result in undesired effects on the crude oil production. Burning of associated gas has become an undesired option in view of depletion of hydrocarbon sources and air pollution. Gas found together with crude oil is known as associated gas, whereas gas found separate from crude oil is known as non-associated gas or natural gas. Associated gas may be found as "solution gas" dissolved within the crude oil, and/or as "gas cap gas" adjacent to the main layer of crude oil. Associated gas is usually much richer in the larger hydrocarbon molecules (ethane, propane, butane) than non-associated gas.--

Paragraph at line 31 of page 2, ending at line 5 of page 3, has been amended as follows:

Paragraph at line 5 of page 2 has been amended as follows:

-- Numerous types of reactor systems have been used for carrying out the Fischer Tropsch reaction. The developed Fischer Tropsch reactor systems include fixed bed reactors, especially multitubular fixed bed reactors, ~~fluidised~~ fluidized bed reactors as entrained ~~fluidised~~ fluidized bed reactors and fixed ~~fluidised~~ fluidized bed reactors, and slurry bed reactors as three phase slurry bubble columns and ebulated bed reactors. --

Paragraph at line 5 of page 4 has been amended as follows:

-- The Fischer Tropsch reaction is characterized by a very high heat of reaction. Unfortunately, the heat transfer characteristics of fixed-bed reactors are generally poor because of the relatively low mass velocity. If one attempts, however, to improve the heat transfer by increasing the gas velocity, a higher CO conversion can be obtained but there is an excessive pressure drop across the reactor, which limits commercial viability. In order to obtain the CO conversions desired and gas throughputs of commercial interest, the needed conditions result in a high radial temperature profile. For that reason, the Fischer-Tropsch fixed-bed reactor diameter should be less than 5 or 7 cm to avoid these excessive radial temperature profiles. The desired use of high-activity catalysts in Fischer-Tropsch fixed-bed reactors, makes the situation even more worse undesirable. The poor heat transfer characteristics makes local run aways possible (hot spots), which may result in local deactivation of the catalyst. Often an axial temperature profile ~~exist~~ exists over the tube. As a certain maximum temperature cannot be exceeded, part of the catalyst works at a sub-optimum level.--

Paragraph at line 27 of page 4, ending at line 7 of page 5, has been amended as follows:

-- As indicated above, the use of catalyst particle sizes greater than 200 micron diameter to avoid excessive pressure drop through the reactor results in high methane selectivity and low selectivities toward the high molecular weight paraffins, which generally have more economic value. This selectivity is due to a disproportional catalyst pore diffusion limitation on the rate of transport of reactants (CO and H₂) into the interior of the catalyst particle. To address the situation, the use of catalysts particles having the active metal component restricted to a thin layer on the outer edge of the particle has been suggested. These catalysts appear costly to prepare and do not appear to make good use of the available reactor volume.--

Paragraph at line 8 of page 5 has been amended as follows:

-- The use of liquid recycles as a means of improving the overall performance in a fixed-bed design has been described. Such a system is also called a "trickle bed" reactor (as part of a subset of fixed-bed reactor systems) in which both reactant gas and an inert liquid are introduced (preferably in an upflow or down flow orientation with respect to the catalyst) simultaneously. The presence of the flowing reactant gas and liquid improves the reactor performance with respect to CO conversion and product selectivity. A limitation of the trickle bed system (as well as of any fixed-bed design) is the pressure drop associated with operating at high mass velocities. The gas-filled voidage in fixed-beds (typically <0.50) does not permit high mass velocities without excessive pressure drops. ~~A too high a~~ A pressure drop that is too large can cause particle attrition/crushing. Consequently, the mass throughput undergoing conversion per unit reactor volume is limited due to the heat transfer rates. Increasing the individual catalyst particle size may slightly improve heat transfer by allowing higher mass velocities (for a given pressure drop), but the loss in selectivity toward the high boiling point products and the increase in methane selectivity combined with the increase in catalyst activity generally offset the commercial incentives of higher heat transfer.--

Paragraph at line 30 of page 7, ending at line 4 of page 8, has been amended as follows:

-- Small particles can be used in these systems because they are readily ~~fluidised~~ fluidized by the gas flow. The pressure drop across the reactor is limited to approximately the static head of the bed. Small particles, because of their large surface area also result in improved liquid-solid mass transfer compared to fixed-bed Fischer-Tropsch hydrocarbon synthesis reactors. Ultimately, the particle size is limited by the solids management system.--

Paragraph at line 1 of page 9 has been amended as follows:

-- ~~Fluidised~~ Fluidized bed type Fischer-Tropsch reactors also give much better heat transfer characteristics than fixed bed reactors and can employ very small catalyst particles. These reactors operate essentially "dry", which means that the production rates of species which are liquid at reactor conditions must be very low, approaching zero. Otherwise, rapid catalyst defluidization can occur. In practice, this requires very high reactor operating temperatures, which typically lead to high selectivities to methane and the production of a number of less desirable chemical species, such as aromatics.

Catalyst/gas separation can also be a significant technical and economic hurdle with ~~fluidised~~-fluidized bed systems.--

Paragraph at line 29 of page 9, ending at line 8 of page 10, has been amended as follows:

-- ~~An object of the present invention is~~ It would be useful to provide an efficient, low cost, compact process scheme to overcome the disadvantages of the above described processes for the production of especially normally liquid hydrocarbons from gaseous hydrocarbonaceous feedstocks. More especially ~~the process of the invention relates~~ it would be useful to a process which converts the feedstock in a very high selectivity into the desired hydrocarbons. ~~Associated with the very high selectivity, a very high thermal efficiency is obtained. Using the process of the invention, C₅+ carbon efficiencies of more than 90% can be obtained, while thermal efficiencies, for a fully optimised process, of above 75% can be obtained.~~ --

On page 10, delete line 9-26.

On page 10, above line 27, insert--Summary of the Invention--

Paragraph at line 27 of page 10 has been amended as follows:

-- The present invention ~~therefore relates~~ is directed to a process for the preparation of hydrocarbons and the generation of heat by reaction of carbon monoxide and hydrogen in the presence of a catalyst at elevated temperature and pressure in at least two stages, the process comprising:

- i) introducing a gas comprising carbon monoxide and hydrogen into a first reactor section comprising catalyst and introducing cooling ~~fluidum~~ fluid into this first reactor section;
- ii) allowing a part of the carbon monoxide and hydrogen to react catalytically in the first reactor section to hydrocarbons and water, at least part of the reaction heat being absorbed directly by the cooling ~~fluidum~~ fluid;
- iii) withdrawing from the reactor section a stream consisting of the reaction product comprising the hydrocarbons, water, unconverted feed and cooling ~~fluidum~~ fluid;
- iv) cooling down at least part of the withdrawn stream comprising cooling ~~fluidum~~ fluid to generate heat;

- v) optionally removing water from the withdrawn stream;
- vi) introducing stream obtained in step v) comprising at least unconverted carbon monoxide and hydrogen into a second or further reactor section comprising catalyst and introducing cooling fluidum fluid into this second or further reactor section;
- vii) optionally introducing a hydrogen containing stream into the second or further reactor section;
- viii) allowing a part of the carbon monoxide and hydrogen to react catalytically in the second or further reactor section to hydrocarbons and water, at least part of the reaction heat being absorbed directly by the cooling fluidum fluid;
- ix) optionally repeating steps iii-viii further reactor sections; and,
- x) withdrawing from the last reactor section the reaction product comprising the hydrocarbons, water, any unconverted carbon monoxide, any unconverted hydrogen and cooling fluidum fluid.

On page 12, above line 1 insert--Detailed Description of the Invention--

In the present invention, the Fischer Tropsch reaction may be carried out in two or more, preferably adiabatic, reactor sections, each reactor section comprising a, preferably high voidage, fixed catalyst bed, in which reactants and cooling medium may be introduced into the reactor sections, the reactants being partly converted, the cooling medium directly absorbing the heat generated in the Fischer Tropsch reaction. The reaction products, unconverted feed and heated cooling medium may be withdrawn from the reactor sections, unconverted feed may be, at least partly, reintroduced into (another) one of the reactor sections, hydrocarbon products may be withdrawn, water formed in the Fischer Tropsch reaction may preferably be removed and heated cooling medium may be cooled down under the simultaneous generation of heat and reintroduced into the reactor sections. Preferably hydrogen is added to the reactants between the reactor sections.

In association with a very high selectivity, a very high thermal efficiency may be obtained. Using the process of the invention, C5+ carbon selectivities of more than 90% can be obtained, while thermal efficiencies, for a fully optimized process, of above 75% may be obtained.--

Paragraph at line 1 of page 12 has been amended as follows:

-- ~~An important advantage~~ One aspect of the proposed inventive process is the possibility to reach very high CO conversion levels and very high C₅+ selectivities.

Further, when compared with the usual fixed bed reactors, a product is obtained in which the amount of olefins is relatively high. This makes the product more useful for chemical applications. The relatively low pressure drop avoids the use of a large (and expensive) compressor. No gas recycle is needed to obtain the high conversion. The scale up of the fixed bed reactor is relatively easy. Catalyst loading and unloading is fairly simple when compared with the traditional fixed bed reactor. Introduction of structured catalysts, e.g. monolithic structures or plate structures covered with a thin layer of catalyst ~~can~~ may easily be done. Optimum use ~~can~~ may be made of the catalyst in view of the relatively short reactor beds, resulting in relatively flat temperature profiles. The use of the cooling ~~fluidum~~ fluid results in much improved heat transfer characteristics when compared with traditional fixed bed reactors. The use of a number of reactor sections makes it possible to adapt the total process in several ways;. ~~For example, for instance~~ different catalysts ~~can~~ may be used in different reactor sections, while the temperature of each reactor sections ~~can~~ may be controlled ~~in an independent way~~ independently. Further, the catalyst may differ in size in each reactor section to use the total reactor space as ~~efficient~~ efficiently as possible. The removal of water between the stages allows higher reactant partial pressures (at the same total pressure), and results in less carbon dioxide formation. The potential addition of hydrogen between the stages makes it possible to use low H₂/CO ratios, resulting in high selectivities to C₅+ hydrocarbons. The absence of cooling internals in the reactor sections makes the construction of the reactor relatively easy and relatively cheap. Furthermore, no expensive reactor space is needed for an indirect cooling system. Standard heat exchange equipment ~~can~~ may be used for cooling down the cooling ~~fluidum~~ fluid. No expensive tube sheets are necessary. The problems encountered with slurry systems such as scale up, kinetics control, back mixing, gas distribution and solids management do not exist. Much higher conversion and selectivities are obtained.—

Paragraph at line 10 of page 13 has been amended as follows:

-- The number of stages (or reactor sections) is at least 2, preferably at least 3 in order to obtain a minimum of the above ~~the~~ described advantages. The maximum number

may be up to 50 or even higher, but in order not to excessively complicate ~~to make~~ the process (and all hardware involved) and process control ~~not to complicated~~, a number of at most 40 stages is preferred. Very suitably, to combine the optimum advantages of the new process and a not too complicated process and process control, the number of stages is between 5 and 20, more preferably between 8 and 12. In principle, each reactor section can may be operated in one reactor. It is preferred to combine several sections in one reactor. Suitably at least 2 sections are combined in one reactor, while at most 25 reactor sections, preferably at most 15, are combined in one reactor. Too many reactor sections will result in more complicated hardware and process control. More preferably between 3 and 7 sections are combined in one reactor.--

Paragraph at line 28 of page 13 has been amended as follows:

-- The H₂/CO (molar) ratio of the feed gas to the first reactor section may be between ~~3~~ 0.3 and ~~0.3~~ 3 or higher or lower. Very ~~suitable~~ suitably the H₂/CO ratio is between ~~0.4~~ 2-0 and ~~2.0~~ 0-4, especially between ~~0.4~~ 1-6 and ~~1.6~~ 0-4, preferably between ~~0.5~~ 1-1 and ~~1.1~~ 0-5. It will be appreciated that lower H₂/CO ratios result in higher C₅+ selectivities. Thus low ratios are preferred. As the consumption ratio is usually between 2.0 and 2.1, the use of a feed ratio below the consumption ~~ratio~~ ratio will result in a decrease of the H₂/CO ratio during the reaction. ~~It is desired that~~ Preferably, the ratio ~~does~~ should not fall below 0.2, in order to avoid undesired side reactions, especially the formation of coke on the catalyst. In a preferred embodiment, the feed ratio to each reactor section is below the consumption-~~ratio~~ ratio, e.g. between 0.5 ~~1-1~~ and 1.1 ~~0-5~~, and hydrogen is added between the stages to increase the ratio again to a higher value, preferably to a value between ~~0.4~~ 1-6 and ~~1.6~~ 0-4, more preferably between ~~0.5~~ 1-1 and ~~1.1~~ 0-5. Hydrogen is preferably added as substantially pure hydrogen (i.e. more than 98 vol% hydrogen). However, ~~also~~ synthesis gas having a (very) high H₂/CO ratio may ~~also~~ be used. For instance, a ratio of 4 may be used, preferably 6, more preferably 10.--

Paragraph at line 21 of page 14 has been amended as follows:

-- The CO conversion per stage is may suitably be between 2 and 50 vol%, preferably between 3 and 40 vol%, more preferably between 6 and 15 vol% (conversion of CO based on feed stream to the first reactor section). It will be appreciated that the

conversion per stage will be related to the total number of reactor section. For instance, at a number of sections between 8 and 12, the CO conversion per stage will be between 8.3 ~~12.5~~ and 12.5 ~~8.3~~ vol%.--

Paragraph at line 29 of page 14, ending at line 7 of page 15, has been amended as follows:

--The process of the invention is may suitably be carried out in such a way that in the first reactor section, preferably all reactor sections, at least 50%, especially at least 80%, of the heat generated by the reaction is directly absorbed by the cooling ~~fluidum~~ fluid, preferably at least 90%, more preferably at least 95%. Part of the heat may be removed by indirect cooling by means of a cooling system in the reactor section. This, however, is not a preferred embodiment. Additional indirect cooling may be used in a particular part of the section, in order to suppress the temperature locally, e.g. to avoid a particular maximum in the heat profile over the section.--

Paragraph at line 18 of page 15 has been amended as follows:

-- The temperature increase of the cooling fluid per reactor section is suitably between 3 and 30 °C, preferably between 5 and 20 °C, more preferably between 7 and 15 °C. At lower levels the process ~~will~~ may be less efficient, at higher levels the temperature difference between the entrance and the end of the catalyst bed ~~will~~ may become too high. A ~~too-high~~ temperature that is too high at the end may result in a decrease of C₅+ selectivity, and in some cases even catalyst deactivation, a ~~too-low~~ temperature that is too low at the entrance of the bed results in less efficient use of the catalyst.--

Paragraph at line 29 of page 15, ending at line 4 of page 16, has been amended as follows:

--The process of the invention is may suitably be carried out at a GHSV of the carbon monoxide and the total hydrogen together between 2000 and 20000 NI/l/h, preferably between 3000 and 10000 NI/l/h based on total catalyst volume (including voids). The above feed stream comprises the feed to the first reactor section, as well as the intermediate hydrogen additions, inclusive of any carbon monoxide when present. It does not comprise any inerts (methane, nitrogen, steam, etc.).--

Paragraph at line 5 of page 16 has been amended as follows:

-- The process according to the invention usually uses a volume ratio (STP) between the gas fraction and the cooling ~~fluidum~~ fluid fraction introduced in each reactor section is between 0.3 and 3, preferably 0.5 and 2, more preferably about 1. A lower value results in insufficient cooling capacity, a higher ratio will result in a too much ~~large amount~~ of cooling fluid, which makes the reaction less efficient.--

Paragraph at line 13 of page 16, ending at line 6 of page 17, has been amended as follows:

-- The catalyst to be used in the present process ~~comprises suitably~~ suitably comprise one or more metals active in the Fischer Tropsch reaction. Very suitable are iron, cobalt or nickel on a carrier, especially cobalt, preferably in combination with one or more promoters. The amount of catalytically active metal on the carrier (calculated as pure metal) is preferably in the range of from 3 to 300 pbw per 100 pbw of carrier material, more preferably from 10 to 80 pbw, especially from 20 to 60 pbw. The promoters may be selected from one or more metals or metal oxides. Suitable metal oxide promoters may be selected from Groups IIA, IIIB, IVB, VB and VIB of the Periodic Table of Elements, or the actinides and lanthanides. In particular, oxides of magnesium, calcium, strontium, barium, scandium, yttrium, lanthanum, cerium, titanium, zirconium, hafnium, thorium, uranium, vanadium, chromium and manganese are most suitable promoters. Particularly preferred metal oxide promoters for the catalyst used to prepare the waxes for use in the present invention are manganese and zirconium oxide. Suitable metal promoters may be selected from Groups VIIB or VIII of the Periodic Table. Rhenium and Group VIII noble metals are particularly suitable, with platinum and palladium being especially preferred. The amount of promoter present in the catalyst is suitably in the range of from 0.01 to 100 pbw, preferably 0.1 to 40, more preferably 1 to 20 pbw, per 100 pbw of carrier.--

Paragraph at line 7 of page 17 has been amended as follows:

-- The process of the invention suitably uses a catalyst system in the form of a fixed bed, preferably a fixed bed having a void volume between 50 and 85 vol%, preferably between 60 and 80 vol%. In principle any shape of the catalyst is possible. Spheres, hollow spheres, extrudates, hollow extrudates, rings, saddles, structured packings etc. are possible. In order to reach the preferred void volumes, the fixed bed comprises

preferably one or more monolithic structures, preferably ceramic monolithic structures, metal extruded ~~monolithes~~ monoliths or carbon ~~monolithes~~ monoliths, layers of corrugated plates, especially metal corrugated plates, gauzes, especially metal gauzes or shavings, especially metal shavings. The ceramic carrier is suitably a porous refractory oxide, preferably selected from silica, alumina, titania, zirconia. In another embodiment the carrier is a plate, gauze or shaving made from ~~aluminium~~ aluminum, iron or copper, especially stainless steel. It will be appreciated that the all reactor sections may comprise the same catalyst, but also that different reactor sections may contain different catalysts. Depending on the exact composition of the feed for a certain reactor section and the objective to be met by the specific reactor section, a different catalyst may be used. In addition, depending on feed, catalyst and objective specific reaction conditions may be used in the reactor sections.

Paragraph at line 1 of page 18 has been amended as follows:

-- The cooling ~~fluidum~~ fluid to be used in the process of the invention ~~suitably consists of~~ may be one or more organic compounds, preferably Fischer Tropsch hydrocarbons, more especially C_{14+} Fischer Tropsch hydrocarbons. It will be appreciated that at the start of the reaction a certain ~~fluidum~~ fluid may be used, however, when the cooling ~~fluidum~~ fluid is used in a recirculating process, which is a preferred embodiment, the starting cooling ~~fluidum~~ fluid will be removed from the reaction together with the liquid reaction product, and gradually the cooling ~~fluidum~~ fluid will be replaced by Fischer Tropsch liquid product. It will be appreciated that the cooling ~~fluidum~~ fluid is preferably inert and stable during the reaction conditions.--

Paragraph at line 14 of page 18 has been amended as follows:

-- In the present process, heat is exchanged in such a way that the temperature of the stream cooling ~~fluidum~~ fluid withdrawn from any reactor section and to be introduced in another section is decreased by 5-20 °C, preferably 7-15 °C, more preferably by the temperature increase of the reactor section involved. In that way a stable process is obtained. In specific circumstances, the amount of heat exchange is adjusted in such a way that a temperature profile is created over the all reactor sections, preferably a continuous temperature increase over all reactor sections. Also depending on specific catalysts more or less heat may be exchanged in order to create the desired temperature in each reactor section.--

Paragraph at line 27 of page 18, ending at line 10 of page 20, has been amended as follows:

-- Suitably, the stream withdrawn from a reactor section is separated into a liquid stream and a gaseous stream, followed by cooling down the liquid stream and cooling down the gaseous stream, suitably to a temperature between 80 and 150 °C, preferably to a temperature between 90 and 130 °C. The liquid stream comprises liquid reaction product and cooling ~~fluidum~~ fluid, the gaseous product comprises unconverted reactants, gaseous hydrocarbon products, steam and, if present, inerts. It will be appreciated that it is also possible to first cool down the stream withdrawn from a reactor section followed by separation into a liquid stream and a gaseous stream, followed by cooling down the gaseous stream. Also Other combinations are also possible. The cooled down liquid stream is used as cooling ~~fluidum~~ fluid in the same or another reactor section. Part of the cooled down product comprising the liquid product is to be removed from the process as the desired product, or is sent to a further work-up section. As, in most cases, the cooling ~~fluidum~~ fluid will be the same as the reaction product, there is no need to separate between cooling ~~fluidum~~ fluid and reaction product. The gaseous stream is cooled down suitably to a temperature between 80 and 150 °C, preferably to a temperature between 90 and 130 °C. Cooling down the gaseous stream results in the condensation of hydrocarbons and water. The water is preferably separated from the condensation product. The hydrocarbon stream will leave the process as the desired product, or is sent to a further work-up section. It is less desired to use the condensed product as cooling ~~fluidum~~ fluid in one or more reactor sections, as a large amount will evaporate under the reaction conditions, resulting in a decrease of reactant partial pressures. The remaining gaseous stream, comprising at least unconverted feed, is introduced into the following reactor section. ~~Preferable~~ Preferably additional hydrogen is used ~~to in~~ this stream. For efficiency ~~reason~~ reasons, ~~there it is the possibility possible~~ to combine the gaseous streams from two or more reactors, followed by cooling down. ~~Please note that it~~ It is also possible to use two or more equivalent reactors comprising equivalent reactor sections, and to combine withdrawn streams from equivalent reactor section, followed by combined cooling and further processing, after which the gaseous streams and cooling ~~fluidum~~ fluid are reintroduced in the following equivalent reactor sections. Similar constructions are also possible for the cooling down of the liquid streams, as well in one reactor comprising several reactor sections, as well as between two

or more equivalent reactors. Part or all water may also be removed by means of membrane separation from the withdrawn gaseous streams.--

Paragraph at line 11 of page 20 has been amended as follows:

-- Preferably the amount of water which is removed from the withdrawn stream after a reactor section is between 50 and 95% of the water formed in the reaction, preferably between 60 and 90%. This ~~can~~ may be obtained by using the preferred temperature ranges as described above.--

Paragraph at line 17 of page 20 has been amended as follows:

-- It will be appreciated that cooled down cooling ~~fluidum~~ fluid from a reactor section may be introduced into the same reactor section or into a different reactor section. Suitably the cooled down cooling ~~fluidum~~ fluid from a reactor section is introduced into the next reactor section. Further, cooling ~~fluidum~~ fluid from a number of reaction sections may combined and re-introduced in a number of reactor sections. As the cooling ~~fluidum~~ fluid absorbs most of the heat generated in a reactor section, it will be clear that the temperature control of a particular section ~~can~~ may be ~~realised~~ realized by the amount of cooling ~~fluidum~~ fluid sent to a particular reactor section and by the temperature of the cooling ~~fluidum~~ fluid. The preferred option is the control of the amount, as this can be easily controlled.--

Paragraph at line 11 of page 21 has been amended as follows:

-- The process of the present invention is suitably carried out with a mixture of hydrogen and carbon monoxide without any inert gases. This results in the most efficient process. As, however, the pressure drop, when compared with the usual fixed bed reactors is considerable less severe, it is also possible to use synthesis gas containing a certain amount of inerts. Suitably the gas feed to the first reactor section may comprises up ~~till~~ to 50 vol% inerts, preferably up ~~till~~ to 20 vol%, more preferably up till 10 vol%. The inerts, especially nitrogen, may be present in the oxygen containing gas stream which is used in the partial oxidation of the hydrocarbonaceous feed, or may be present in the hydrocarbonaceous feed itself, for instance nitrogen and/or noble gases in natural gas.--

Paragraph at line 26 of page 21 has been amended as follows:

-- The normally liquid hydrocarbons are especially mixtures of C₅-C₁₈ hydrocarbons, although small amounts of C₄- and C₁₉+ compounds may be present. At STP, these mixtures are liquid. C₁-C₄ compounds are considered as normally gaseous hydrocarbons. Normally solid hydrocarbons are especially mixtures of C₁₉+ compounds, up to C₂₀₀. Smaller quantities of C₁₈- may be present. Normally solid hydrocarbons are solid at STP. The hydrocarbon mixture made in the Fischer Tropsch process vary from C₁ to C₂₀₀ or even higher. The amount of C₁₉+ hydrocarbons is preferably at least 60 wt%, preferably 70 wt%, more preferably 80 wt%. These hydrocarbons are paraffinic in nature, although considerable amounts of olefins and/or oxygenates may be present. Suitably up to 20 wt%, preferably up to 10 wt%, of either olefins or oxygenated compounds may be present. The compounds are mostly normal compounds, although a few wt% of branched, especially methyl branched, may be present.--

Paragraph at line 24 of page 22 has been amended as follows:

-- The higher boiling range paraffinic hydrocarbons obtained in the present process may be isolated and subjected to a catalytic hydrocracking, which is known per se in the art, to yield middle distillates. The catalytic hydrocracking is may be carried out by contacting the paraffinic hydrocarbons at elevated temperature and pressure and in the presence of hydrogen with a catalyst containing one or more metals having hydrogenation activity, and supported on a carrier. Suitable hydrocracking catalysts include catalysts comprising metals selected from Groups VIB and VIII of the Periodic Table of Elements. Preferably, the hydrocracking catalysts contain one or more noble metals from group VIII. Preferred noble metals are platinum, palladium, rhodium, ruthenium, iridium and osmium. Most preferred catalysts for use in the hydrocracking stage are those comprising platinum. To keep the process as simple as possible, the hydrocracking will usually not be a preferred option.--

Paragraph at line 31 of page 23 has been amended as follows:

-- The invention further relates to one or more reactors for carrying out the process as described above. A very suitable reactor is an elongated cylindrical vessel, which, when in use, ~~will~~may be a vertical reactor. In one of the preferred embodiments, in which one reactor comprises 3 to 7 reactor sections, the reactor will contain 2 to 6 plates,

suitably at about the same distance, thus creating the 3 to 7 reactor sections. Also when in use, the 2 to 6 plates dividing the reactor in the several reactor sections, are preferably in a horizontal position. Each reactor section will contain a fixed catalyst bed, means for distributing gas and liquid over the catalyst bed at the upstream end of the catalyst bed and means for collecting gas and liquid at the downstream end of the catalyst bed. In a one suitable design only one (large) catalyst bed will be present, delimited by the outside reactor wall. Above the catalyst bed will be a space for the distribution means, below the reactor bed there will be a space for collecting gas and liquid. Gas and liquid may be removed from the reactor section by one or more pipes for the liquid, and one or more pipes for the gas. In an alternative, gas and liquid may be removed via one or more common pipes, followed by separation (in one or more standard separation vessels) outside the reactor. Usually gas and cooling ~~fluidum~~ fluid will be introduced into the top of the first reactor section above the catalyst bed. Gas and liquid will be removed from the reactor at the lower end of the first section, and, after separation, cooling, removal of liquid product and, often, water removal and optional addition of hydrogen, introduced into the top of the second section etc. As discussed above removal of water may be done after each reactor section, but also after each second, or even third, reactor section. Also, the liquid streams of several sections may be combined and cooled, followed by reintroduction into the reactor sections. Liquid from one section ~~may be introduced~~ may also be reintroduced into the same reactor section. The gas stream will in most cases flow from the first section to the second section, to the third section etc. Beside vertical reactors, it is also possible to use horizontal reactors. These horizontal reactors may comprise similar compartments as described for the vertical reactors, but may also comprise compartments with structured catalyst packings containing substantially horizontal channels through which a gas/liquid dispersion is transferred in a horizontal direction.--

On page 26 , above line 1, insert --We claim:--